



Activity and stability of powder and monolith-coated Ni/GDC catalysts for CO₂ methanation

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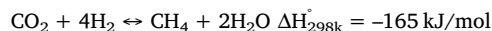
ABSTRACT

The methanation of CO₂ via the Sabatier process is gaining interest for power-to-gas (P2G) application. In this work, CO₂ methanation activity and stability were investigated over Ni/GDC (gadolinium-doped-ceria) catalysts at atmospheric pressure varying reaction temperature ($T_{\text{SET}} = 300\text{--}600\text{ }^{\circ}\text{C}$) and space velocity ($\text{GHSV} = 10,000\text{--}50,000\text{ h}^{-1}$). Powder catalysts with different Ni content (15–50 wt.%) were synthesized by the solution combustion synthesis (SCS). The same method was adopted to *in situ* deposit the Ni/GDC (50 wt.%Ni) coating layer on the cordierite monolith (500 cpsi). The catalysts were characterized by N₂ adsorption-desorption, X-ray diffraction (XRD), H₂ temperature programmed reduction (H₂-TPR), CO₂ temperature programmed desorption (CO₂-TPD), X-ray photoelectron spectroscopy (XPS), transmission electron microscopy (TEM) and scanning electron microscopy (SEM). Temperature profiles along the structured catalytic bed were discussed to interpret the experimental results.

Catalytic performance increased by increasing the Ni content due to enhanced metal-to-support interaction, basicity and oxygen vacancies. Uniform, thin and high-resistance catalytic layers were *in situ* deposited on the cordierite monoliths by the fully reproducible SCS method. Structured catalysts showed high methane productivity per unit weight of catalyst due to simultaneous low catalytic loading and high flow rate. Excellent stability was observed over 200 h of time-on-stream. The results reported in this manuscript pinpointed on the important aspects of realizing CO₂ methanation on structured catalysts, providing a platform for further optimization studies.

1. Introduction

A successful transition towards a cleaner and more sustainable energy system in 2050 requires large-scale implementation of sustainable and renewable energy source. In contrast with conventional energy sources, the intermittency and fluctuation of renewable energy make difficult its integration into the existing energy grid [1,2]. Recently, Power-to-Gas (P2G) technologies have been introduced with the aim to store the excess of non-programmable renewable energy. In particular, P2G technologies involve two stage: (i) conversion of electrical power to hydrogen via electrolysis and (ii) conversion of hydrogen to high-value chemicals or fuels thought reaction with CO₂ [3–5]. In this context, the production of Substitute Natural Gas (SNG) by CO₂ methanation, also called Sabatier reaction, proceeds according to [6,7]:



It is a reaction of great technological and environmental potential, leading to (i) storage of excess H₂ generated from renewable energy, (ii) reduction of CO₂ emissions (greenhouse gas) from the atmosphere and (iii) production of SNG whose distribution infrastructures are readily available [8–10].

CO₂ methanation is a complex reaction because it is strongly exothermic and thermodynamically favored at low temperatures where kinetic rates are low [11]. Moreover, the high concentration of CO₂ involved results in large potential temperature increases and hot spot formation, lowering the yield and leading to catalyst deactivation by sintering and carbon deposition. For these reasons, the catalyst should be active and stable both at low and high temperature [12]. In the last three decades, several systems have been widely investigated as catalyst

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for CO₂ methanation. Between various transition metals (Rh, Ru, Pd, Pt, Fe, Co), Ni-based catalysts seem to be a good choice due to relatively high intrinsic activity and low price [13–15]. Several supports, including Al₂O₃, ZrO₂, SiO₂, TiO₂, CeO₂, MgO and zeolites, were utilized for supported Ni catalysts [16–22]. Moreover, the catalytic performance of Ni catalysts depends on several factors, such as morphology, size and dispersion, oxygen vacancies, ionic conductivity, metal-support interactions, thermal and mechanical stability [23]. These properties depend in turn on the Ni content, the type of support and promoters and the synthesis method [10,24]. Le and co-workers [16] compared the catalytic performance for CO₂ methanation over Ni/CeO₂ catalysts with different Ni contents, whose activity increased by increasing the nickel load, reaching the maximum with the Ni_{0.8}Ce_{0.2}O_x system. Tada et al. [25], studied CO₂ methanation performances over Ni-supported catalysts evaluating the effect of electronic interactions between metal phase and support. The participation of CeO₂ effectively promoted the activation of CO₂, improving the CO₂ methanation performances [25]. Moreover, the addition of Gd₂O₃ as promoter can modify structural properties, mainly crystallite size and surface area, while simultaneously improving thermal and mechanical resistance of the system [26,27].

The methanation reaction is predominantly carried out in fixed-bed reactors [28]. A structured reactor would be a candidate reaction system providing many advantages such as high surface-to-volume ratio, low pressure drop, good interphase mass transfer and intensification of mass and heat transfer [29–31]. In particular, the fast heat removal from the reactor system allows (i) avoiding limitations in methane yield due to approaching the chemical equilibrium which is a disfavored with increasing temperature and (ii) minimizing catalyst deactivation (metal sintering and carbon deposition) due to thermal stress [32,33]. Fukuhara et al. [32] prepared honeycomb-type Ni/CeO₂ catalysts with high and steady catalytic performance, mainly due to enhanced heat and mass transfer properties [32]. Similar results were reported by Danaci and co-workers [8], studying the CO₂ methanation over macro-porous metal structures coated with Ni/Al₂O₃ catalyst. Despite the benefits of structured systems, the distribution of the catalytic phase on the surface of the monolithic support is a key factor for the overall system performance. Thus, the coated layer needs to be resistant to thermal and mechanical stresses occurring during methanation reaction [34–36]. As previously reported, the solution combustion synthesis (SCS) method is a suitable procedure to deposit uniform, thin and high-strength catalytic layers on the ceramic monoliths [37–41].

In this work, the catalytic properties of Ni/GDC (gadolinium-doped ceria) systems in CO₂ hydrogenation were studied. Powder catalysts with different Ni content (15–50 wt.%) were synthesized by the solution combustion synthesis (SCS). Analogously, structured catalysts were prepared by the same method, *in situ* depositing the Ni/GDC (50 wt.% Ni) coating layer on the cordierite monolith (500 cpsi). The catalysts were characterized by N₂ adsorption-desorption, X-ray diffraction (XRD), H₂ temperature programmed reduction (H₂-TPR), CO₂ temperature programmed desorption (CO₂-TPD), X-ray photoelectron spectroscopy (XPS), transmission electron microscopy (TEM) and scanning electron microscopy (SEM). CO₂ methanation (H₂/CO₂ = 4) activity tests were carried out at atmospheric pressure varying temperature (T_{SET} = 300–600 °C) and space velocity (GHSV = 10,000–50,000 h^{−1}). Stability tests over 200 h of time-on-stream were also performed.

2. Experimental

2.1. Catalysts preparation

2.1.1. Powder catalysts

Ni/GDC catalysts and GDC support (Ce:Gd molar ratio of 0.8:0.2) were synthesized by the solution combustion synthesis (SCS) method,

previously described [42,43]. The Ni loading was set to 15, 30 and 50 wt.% and confirmed by chemical analysis (ICP/OES). The corresponding catalysts were denoted as 15Ni/GDC, 30Ni/GDC and 50Ni/GDC. High-purity reagent-grade were used as received: nickel nitrate (Ni(NO₃)₃·6H₂O, Aldrich), cerium nitrate (Ce(NO₃)₃, Aldrich) and gadolinium nitrate (Gd(NO₃)₃·6H₂O, Aldrich) as metal precursors, while urea (CH₄N₂O, Alfa Aesar) as fuel. In a typical experiment, stoichiometric amounts of metal precursors and urea were dissolved in the minimum quantity of distilled water. The amount of fuel was determined according to stoichiometric balance of oxidizing (O) and reducing (F) valences of the components to have an equivalence ratio (O/F) equal to 1 [44]. The obtained solution was introduced into a furnace preheated at 350 °C where the combustion reaction takes place with rapid increase in temperature and gas evolution (N₂, CO₂, H₂O). The resulting powders were calcined at 600 °C in static air for 2 h.

The catalyst (with the sizes of 200–300 μm) was diluted with quartz (with the same sizes) in the 1:1 volume ratio with a total volume of 1.18 cm³ (diameter 1 cm, length 1.5 cm). The total loading, calculated as quantity of catalytic material per unit volume of catalytic bed, was 0.8 g/cm³.

2.1.2. Structured catalysts

Conventional 500 cpsi cordierite monoliths (cylinder form, diameter 1 cm, length 1.5 cm, specific geometric surface 25.2 cm²/cm³, channels density 77.1 ch/cm², channel size 0.8 mm, wall thickness 320 μm,) were used as support for catalytic layer deposition. Supports were washed in water/acetone (50/50 vol.%) ultrasonic bath for 30 min and dried at 120 °C for 2 h. The catalytic layer (50 wt.% Ni/GDC) was *in situ* deposited by SCS [39,40] dipping each monolith in an aqueous solution containing metal precursors and fuel (as described for powder preparation). The wet monoliths were introduced into a muffle furnace preheated at 600 °C for ca. 10 min, where the combustion reaction occurs, and rapidly cooled down to room temperature in few min. The deposition step was repeated several times to reach the total loading of 0.2 g/cm³ (denoted as MO-A) and 0.5 g/cm³ (denoted as MO-B). These quantities corresponded to approximately 25 and 60% of the catalyst employed in the packed bed system. The resulting monoliths were calcined at 600 °C in static air for 2 h. Fig. 1 shows the photographs of the uncoated monolith (a) and the coated MO-B monolith (b).

2.2. Catalysts characterization

Specific surface area was estimated according to the Brunauer-Emmet-Teller (BET) equation from adsorption/desorption isotherms at liquid nitrogen temperature (−196 °C) on Micromeritics ASAP 2020 instrument. Prior to measurements the samples were degassed under high vacuum for 6 h at 300 °C.

X-ray diffraction (XRD) experiments were carried out on a Philips X-Pert 3710 diffractometer equipped with a Cu Kα radiation at 40 kV and 20 mA. The patterns were recorded at a scanning speed of 1.50° min^{−1} over the range 2θ = 20°–75°. The peaks were assigned according to the PCPFWIN database. The CeO₂ and NiO crystallite sizes were calculated using the Scherrer equation based on the main CeO₂ (111) and NiO (200) reflection peaks.

H₂ temperature programmed reduction (H₂-TPR) was carried out in a Micromeritics ChemiSorb 2750 instrument equipped with a thermal conductivity detector (TCD). A continuous flow of 5% H₂/Ar (30 NmL/min) was passed over the catalyst in the temperature range 20–1000 °C at a heating rate of 20 °C/min. Before measurements, all samples were heat treated at 500 °C for 1 h under O₂ flow (30 NmL/min). The hydrogen consumption was determined based on the H₂-TPR patterns of known amounts of CuO.

CO₂ temperature programmed desorption (CO₂-TPD) was carried out on the same apparatus. The sample was heated up to 800 °C under He flow (30 NmL/min), reduced under H₂ flow (30 NmL/min) at 800 °C for 1 h, cooled to room temperature under He flow (30 NmL/min) and

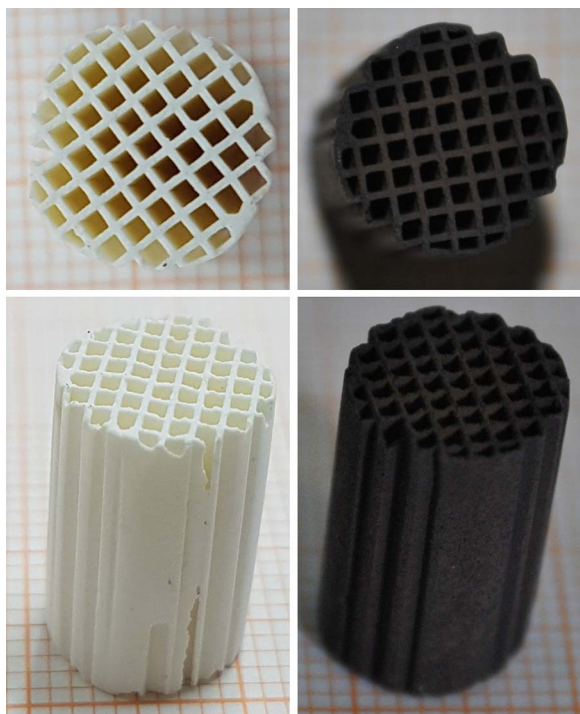


Fig. 1. Photographs of uncoated (left) and coated (right) monoliths.

exposed to CO₂ stream (30 Nml/min) at room temperature for 30 min. The physisorbed CO₂ was removed by He purging at room temperature for 1 h. Then, the temperature was increased with heating rate of 10 °C/min under He flow (30 Nml/min) from 20 to 1000 °C and the desorbed CO₂ was monitored by TCD, which was calibrated by injections of pure CO₂ pulses.

X-ray photoelectron microscopy (XPS) data were collected on a Physical Electronics GMBH PHI 5800-01 spectrometer operating with a monochromatized Al-K_α radiation with a power beam of 300 W. The BE regions of C 1s (275–300 eV), O 1s (520–540 eV) and Ce 3d (875–925 eV) were investigated, taking the C 1s line (284.80 eV) of adventitious carbon as reference. The characterisation experiments were carried out for fresh (after reduction) samples.

Transmission electron microscopy (TEM) micrographs were obtained using a Philips CM12 instrument. The reduced powder samples were dispersed in isopropyl alcohol by ultrasonic treatment and placed on holey copper grids.

The coating procedure was evaluated in terms of homogeneity of the coating, amount of catalyst deposited and adhesion force. Scanning electron microscopy (SEM) images were obtained using a FEI XL 30 equipped with field emission gun. The structured samples were cut longitudinally to evaluate the thickness of the catalytic layer. The adherence of the coating was evaluated in terms of weight loss after two ultrasonic treatment in 50 vol.% isopropyl alcohol solution. The monoliths were treated for 30 min at 45 kHz and 130 W using the USC 900D ultrasonic bath.

2.3. Methanation performance

CO₂ methanation was carried out in a quartz tubular fixed-bed reactor (1 cm inner diameter, 25 cm length) horizontally placed in a furnace under atmospheric pressure. Both powder and structured catalysts were placed in the centre of the quartz tube. The catalyst was heated to 800 °C (15 °C/min) under N₂ flow (30 Nml/min), reduced at 800 °C for 1 h under 50 vol.% H₂/N₂ stream (30 Nml/min) and cooled down to reaction temperature (T_{SET}) under N₂ flow (30 Nml/min) before starting the catalytic test. Four chromel/alumel thermocouples

were employed at a distance of 0.1 (T_{SET}), 0.2 (T_{IN}), 0.5 (T_{CENT}) and 1.4 cm (T_{OUT}) from the beginning of the catalytic bed (length 1.5 cm). The first thermocouple, connected with a PID controller, was used to regulate the reaction temperature (T_{SET}), while the others were used to measure the temperature profile along the catalytic bed. A mixture gas of H₂/CO₂/N₂ with fixed molar ratio 4/1/1 was fed into the reactor by mass flow-controllers (Brooks Instrument Smart Mass Flow). An Agilent 6890 Plus gaschromatograph, equipped with thermal conductivity (TCD) and flame ionization (FID) detectors, was used to on-line analyze reactants and products every 20 min N₂ was used as internal standard for mass balance calibration.

Activity tests (8 h each) were carried out varying temperature (T_{SET} = 300–600 °C) and space velocity (GHSV = 10,000–50,000 h^{−1}). Stability tests over 200 h of time-on-stream were also performed at T_{SET} = 400 °C and GHSV = 30,000 h^{−1}. The experimental results were reported as CO₂ conversion and molar concentrations of H₂, CO, CO₂ and CH₄. Thermodynamic analysis, performed by the simulation package HSC Chemistry[®] 7, involved the determination of equilibrium compositions of the system at specified operating conditions (temperature, pressure). The model was based on the minimization of Gibbs free-energy of the existing species (CH₄, H₂O, CO₂, CO, H₂, solid carbon). This method does not require any detail of chemical reactions taking place in the system while it is based on the concept that at chemical equilibrium the total Gibbs energy of the system has its minimum value [45].

3. Results and discussion

3.1. Powder and structured catalysts characterization

The main textural properties of the synthesized GDC support and Ni/GDC catalysts are reported in Table 1. The surface area ranged from 19.5 to 33.8 m²/g. It could be observed that the incorporation of nickel led to surface area (24.5 for 15Ni/GDC) higher than the GDC support (19.5 m²/g). Moreover, the surface area slightly increased to 33.8 m²/g by increasing the Ni content to 50 wt.% (50Ni/GDC). The diffraction patterns of the calcined samples are displayed in Fig. 2. All the systems showed the XRD peaks for the face-centered cubic phase of CeO₂ (JCPDS Card 4–593), while the absence of the Gd₂O₃ crystallographic planes suggested the amorphous nature of Gd₂O₃ or its presence as Ce_{0.8}Gd_{0.2}O_x solid solution. The CeO₂ average crystallite size calculated from the Scherrer equation, based on the main CeO₂ (111) reflection peak, decreased by increasing the Ni loading (Table 1). As previously reported, the textural features of synthesized powders strongly depend on the flame temperature achieved during the combustion reaction [42]. The theoretical adiabatic flame temperature (T_{ad}) was calculated using the thermodynamic data for reactants and products by the equation:

$$T_{ad} = T_0 \frac{\Delta H_r - \Delta H_p}{C_p} \quad (1)$$

Table 1
Nickel loading, surface area and crystallite size of the synthesized powders.

Sample	Ni content ^a (wt. %)	S _{ABET} (m ² /g)	XRD		
			CeO ₂ Lattice parameter ^b (nm)	CeO ₂ Particle size ^c (nm)	NiO Particle size ^c (nm)
GDC	–	19.5	0.5387	11.3	–
15Ni/GDC	15.1	24.5	0.5378	8.0	6.0
30Ni/GDC	30.4	27.0	0.5378	7.8	7.5
50Ni/GDC	49.8	33.8	0.5377	7.7	9.2

^a Determined by ICP/OES chemical analysis.

^b Calculated by the relation $\alpha = \sqrt{h^2 \cdot k^2 \cdot l^2} \cdot (\lambda / 2 \cdot \sin \theta)$.

^c Calculated by the Sherrer equation of the CeO₂ (111) and NiO (200) reflections.

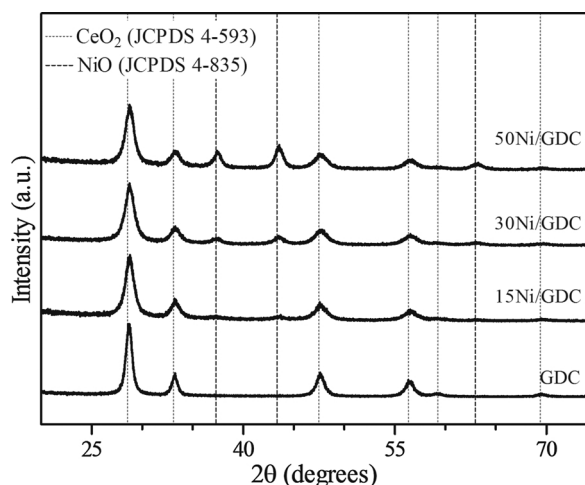


Fig. 2. XRD patterns of the synthesized GDC support and Ni/GDC catalysts.

where ΔH_r and ΔH_p are the formation enthalpies of reactants and products, respectively, C_p is the molar heat capacity of products at constant pressure and T_0 is the ignition temperature. The derived flame temperatures of the prepared samples followed the order: 50Ni/GDC (1250 °C) < 30Ni/GDC (1288 °C) < 15Ni/GDC (1310 °C) < GDC (1318 °C). Indeed, flame temperature decreased by increasing the Ni content, leading to higher surface area and lower CeO_2 particle size values (Table 1). Peaks of NiO (JCPDS 4–835) could be observed in all Ni/GDC catalysts, the intensity of which increased by increasing the Ni content, due to the formation of larger NiO particles. Indeed, the NiO crystallite size slightly increased from 6.0 (15Ni/GDC) to 9.2 (50Ni/GDC). Moreover, the lattice parameter decreased with introducing the Ni metal phase (Table 1). The pure GDC support had a lattice parameter of 0.5387 nm, while the Ni-based systems showed lattice parameters of 0.5377–0.5378 nm, indicating the partial substitution of Ce^{4+} (0.097 nm) with smaller Ni^{2+} (0.081 nm) ions and resulting in the shrinkage of the CeO_2 cell due to the formation of Ce-Gd-Ni-O solid solution.

The reducibility of the combustion-synthesized samples was studied by H_2 -TPR and the results are displayed in Fig. 3. The reduction of bare GDC support showed two main reduction peaks at 550 and 860 °C due to the reduction of CeO_2 surface oxygen and CeO_2 bulk oxygen, respectively [26,46]. The TPR profiles of Ni-based systems showed two main reduction peaks centered at 380 and 450 °C, attributed to the reduction of bulk NiO species with different extent interactions with the

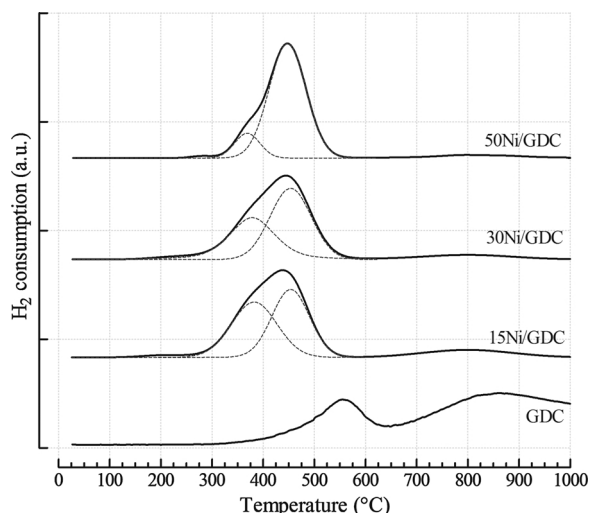


Fig. 3. TPR profiles of synthesized GDC support and Ni/GDC catalysts.

support [14,47,48]. Generally, the reduction of bulk NiO (not reported) occurs in the temperature range of 280–300 °C [49]. The shift to higher temperature of this reduction peak suggested the presence of strong Ni-support interactions [10,49,50]. Moreover, as the Ni loading increased, the high temperature reduction peak became greater in intensity, due to the more pronounced metal-support interaction and to the contribution of larger NiO crystallites, in accordance with XRD evidences (Fig. 2) [51]. Table 2 shows the experimental/theoretical hydrogen consumption and the relative reducibility of the synthesized samples. The experimental H_2 consumption, expressed as mmol_{H_2} per gram of catalyst, was related to the TPR peaks up to 600 °C. In addition, theoretical amount of H_2 demanded was calculated by assuming complete reduction of NiO to Ni for Ni/GDC samples (Table 2). The reducibility of the catalysts decreased with increasing the Ni loading from 91.5% (15Ni/GDC) to 88.4% (30Ni/GDC) and 87.1% (50Ni/GDC). This could be due to the presence of Ce-Gd-Ni-O solid solution, more difficult to reduce, and to the formation of larger NiO particles, as confirmed by XRD analysis. Similar results were reported by Singha et al. [52], indicating that the core of larger Ni oxide particles was not readily exposed to hydrogen for reduction.

Fig. 4 shows the CO_2 -TPD of the Ni/GDC catalysts with various Ni amounts, while the quantitative analysis of the desorbed CO_2 is reported in Table 3. All the catalysts exhibited three main desorption peaks, due to adsorption sites with increasing strength: (a) weak CO_2 adsorption sites situated at low temperatures ($T < 250$ °C); (b) moderate CO_2 adsorption sites at temperature ranging between 250 and 750 °C; (c) strong CO_2 adsorption sites situated at high temperature ($T > 700$ °C) [53]. The desorption profiles in Fig. 4 revealed that the major contribution is due to the presence of moderate-basic sites, following the order 15Ni/GDC ($43.8 \mu\text{mol}_{\text{CO}_2}/\text{g}$) < 30Ni/GDC ($59.7 \mu\text{mol}_{\text{CO}_2}/\text{g}$) < 50Ni/GDC ($75.2 \mu\text{mol}_{\text{CO}_2}/\text{g}$). The same order was observed for the basic site density, expressed as CO_2 adsorbed per unit volume of catalytic bed (Table 3). Besides, weak-basic sites concentration seemed to be not affected by the nickel content, while the amount of medium- and strong-basic sites increased by increasing the Ni loading (Table 3). Thus, the increase in Ni content led to higher basicity mainly due to the stronger Ni-support interaction, as confirmed by TPR analysis [54,55].

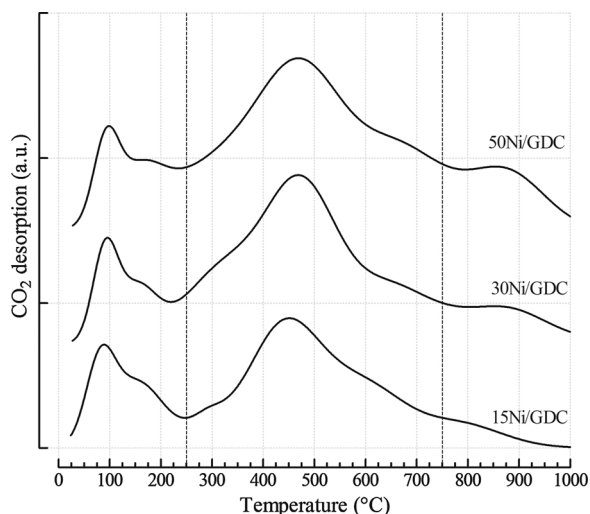
Fig. 5 presents the XPS spectra in the O 1s and Ce 3d regions for the Ni/GDC samples. A nonlinear least-square fitting procedure using Gaussian and Lorentzian type curves showed O 1s spectra with two distinct peaks [56,57]. The intense peak at 528.9 eV could be attributed to the surface lattice oxygen (O_α) of GDC, while the signal at 531.7 eV could be assigned to the surface hydroxyl groups or the adsorbed oxygen (O_β) [58,59]. Based on the area integrals of the oxygen species, the ratio of O_α to O_β ($\text{O}_\beta = \text{O}_\alpha + \text{O}_\beta$) for all catalysts was estimated and summarized in Table 4. It can be seen that the percentage of surface lattice oxygen (O_α) decreased by increasing the Ni content. In particular, the $\text{O}_\alpha/\text{O}_\beta$ ratio decreased from 84.5 to 79.7% for 15Ni/GDC and 30Ni/GDC catalysts, respectively, while a slight decrease to 79.1% was observed further increasing the Ni content (50Ni/GDC). Such behavior could be ascribed to the increasing number of oxygen vacancies caused by the insertion of the Ni^{2+} cations on the GDC support, in accordance with XRD evidences [60,61].

The interpretation of the Ce 3d photoemission spectra is not straightforward mainly due to multielectronic processes both in the Ce^{4+} and Ce^{3+} states [56]. The XPS spectra in Fig. 5 could be deconvoluted into ten peaks, corresponding to four pairs of spin-orbit doublets [23,62]. Generally, the peaks marked as v , v' , v'' , u , u'' and u''' arise from the contribution of Ce^{4+} species, while the peaks labeled as v° , v''' , u° , and u''' originate from the contribution of Ce^{3+} species [62,63]. According to the ratio of Ce^{3+} ion peaks area to that of the total Ce^{3+} and Ce^{4+} ion peaks area, the following equation was used to calculate the relative content of Ce^{3+} :

Table 2

Experimental/theoretical hydrogen consumption and relative reducibility from TPR profiles of the Ni/GDC catalysts.

Sample	Experimental H ₂ consumption (mmol _{H2} /g)			Theoretical H ₂ consumption (mmol _{H2} /g)	Reducibility (%)
	Low temperature peak	High temperature peak	Total		
15Ni/GDC	1.14	1.17	2.31	2.53	91.5
30Ni/GDC	1.78	2.72	4.50	5.09	88.4
50Ni/GDC	0.95	6.31	7.26	8.34	87.1

**Fig. 4.** CO₂-TPD of 15Ni/GDC, 30Ni/GDC and 50Ni/GDC catalysts.**Table 3**CO₂-TPD analysis for determining basic properties of the Ni/GDC catalysts.

Sample	CO ₂ desorption (μmol _{CO2} /g)				Basic site density ^a (μmol _{CO2} /cm ³)
	Weak (< 250 °C)	Moderate (250–750 °C)	Strong (> 750 °C)	Total	
15Ni/GDC	17.1	43.8	4.0	64.9	35.0
30Ni/GDC	16.9	59.7	10.1	86.7	47.8
50Ni/GDC	17.5	75.2	18.2	110.9	60.2

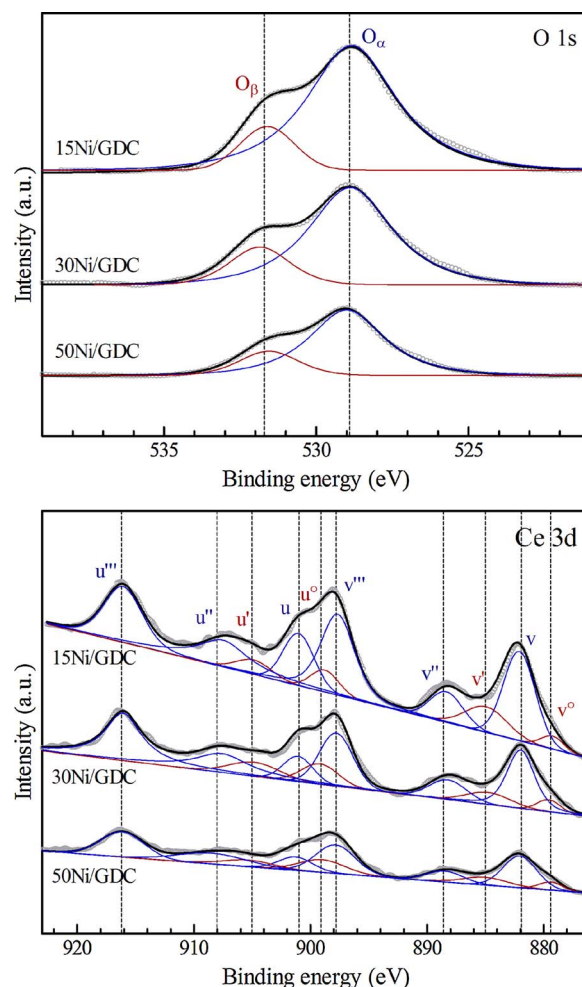
^a Expressed as moderate (250–750 °C) basic sites per unit volume of catalytic bed.

$$\text{Ce}^{3+} = \frac{u'' + u' + v'' + v'}{\sum (u + v)} \quad (2)$$

The corresponding calculation results are summarized in Table 4. The relative content of Ce³⁺ increased by increasing the Ni content, due to the formation of more oxygen vacancies in the ceria lattice. Besides, the 50Ni/GDC catalyst showed the highest Ce³⁺ surface content (21.8%), indicating a more pronounced metal-support interaction, which is consistent with the above TPR results [62,63].

The morphology of 50Ni/GDC catalyst was checked by TEM analysis. TEM images in Fig. 6a showed 30–50 nm CeO₂ particles in agglomerate form. Not very uniform sized Ni particles, ranging between 10 and 25 nm, were revealed. More in detail, Images 5b and 5c show the magnifications of two areas of Fig. 6a; fringes with d-spacing 2.0 Å were attributed to the (111) planes of Ni, whereas fringes with 3.1 Å were attributed to the (111) planes of CeO₂ crystals.

Figs. 7 and 8 show SEM images of bare and coated monoliths. A reduction of the macro-porosity of the bare cordierite (Fig. 7a) was revealed after the in-situ deposition of the catalytic layer (Fig. 7b). Well uniform coating layers were found, as evidenced in the corner (Fig. 7b), inner (Fig. 7c,d) and frontal (Fig. 7e) channel views and from the EDX mapping (Fig. 7f) of the coated MO-B monolith. Moreover, the magnification of the coated layer (Fig. 7d) confirmed the presence of a

**Fig. 5.** High resolution XPS spectra of the Ni/GDC catalysts.**Table 4**

Summary of XPS data for the Ni/GDC catalysts.

Sample	Relative content of O _α (%) ^a	Relative content of Ce ³⁺ (%) ^b
15Ni/GDC	84.5	16.3
30Ni/GDC	79.7	20.1
50Ni/GDC	79.1	21.8

^a Calculated according to the ratio O_α/O_T, based on the area integrals in the O 1s spectra (O_T = O_α + O_β).^b Calculated according to Eq. (2), based on the area integrals in the Ce 3d spectra.

significant residual porosity due to the escaped gas during the SCS steps, as previously reported [40–42]. Average thickness of ca. 10 μm for low-loaded system (MO-A, Fig. 8a) and 40 μm for high-loaded system (MO-B, Fig. 8b) were determined.

The mechanical resistance of the coated layers, evaluated by ultrasonic treatment in isopropyl alcohol solution, evidenced good adherence of the catalytic layer to the cordierite support. Negligible

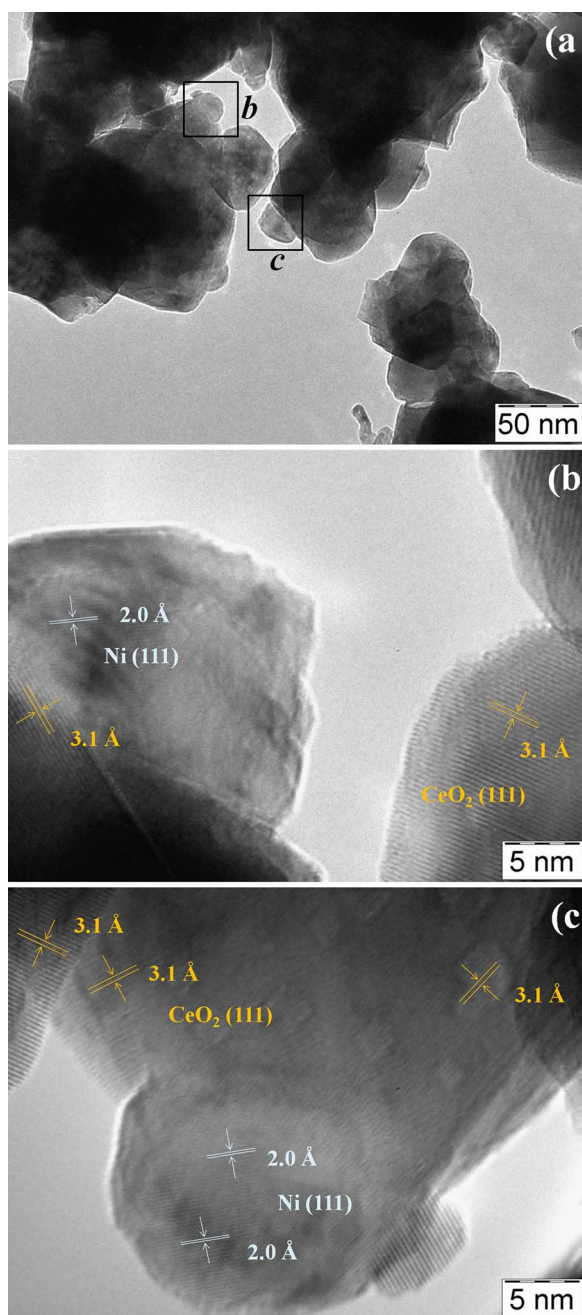


Fig. 6. TEM micrographs of the 50Ni/GDC catalyst.

weight loss of ca. 0.3–0.5% after the second adhesion test (calculated on the total weight of the monolith) were found.

3.2. Catalytic activity

3.2.1. Powder catalyst: effect of Ni content and reaction temperature

Fig. 9 shows the products distribution of CO₂ methanation at equilibrium calculated via the Gibbs free energy minimization method at 1 atm for a feed gas containing H₂ and CO₂ with a stoichiometric H₂/CO₂ molar ratio of 4. At relatively low temperature (200–300 °C), CH₄ and H₂O were the main reaction products. Increasing the temperature, the methanation reaction was disfavoured due to its exothermic nature, leading to increased amounts of unreacted CO₂ and H₂, along with decreased amounts of produced CH₄. Moreover, at temperature higher than 450 °C the reverse water gas shift (RWGS) reaction was favoured, resulting in the increase of CO by-product. However, when the

temperature exceeded 550 °C, the mole fraction of CO₂ reached its maximum and then decreased because the RWGS reaction dominated [64].

The mass transfer in a porous catalyst is a significant factor in determining the catalytic activity [65]. As reported by other Authors, the external diffusion limitation is negligible when the rate of the surface reaction is sufficient slow [66,67]. On the other hand, the internal diffusion limitation was estimated by the Weisz-Prater criteria, according to:

$$N_{WP} = \frac{\mathcal{R} \cdot R_p^2}{C_S \cdot D_{eff}} < 1 \quad (3)$$

where \mathcal{R} is the reaction rate of CO₂ per unit volume of the catalyst (1.48×10^{-5} mol/cm³ s);

R_p is the particle radius of catalyst (1.25×10^{-2} cm);

C_S is the surface concentration of CO₂ (3.02×10^{-6} mol/cm³), assumed equal to the bulk CO₂ concentration and calculated by the equation $C_S = P_{CO_2}/RT$;

D_{eff} is the pore diffusion (3.41×10^{-2} cm²/s), dominated by Knudsen diffusion (D_{Kn}) and calculated by $D_{eff} \approx D_{Kn} = 1/3\lambda d$, where λ is the average molecular velocity (5.69×10^4 cm/s) and d is the average pore diameter (1.80×10^{-6} cm).

Under the selected reaction conditions (i.e. $P=1$ atm, $T=673$ K), the value of N_{WP} for the 50Ni/GDC catalyst was 0.022, indicating a negligible internal mass transfer limitation [67].

CO₂ conversion (Fig. 10a) and effluent composition (Fig. 10b) measured during the catalytic tests carried out at atmospheric pressure are plotted in Fig. 10, as a function of Ni content (15–50 wt.%) and reaction temperature ($T_{SET} = 300$ –600 °C). As comparison, the equilibrium values of the CO₂ conversion were also reported as dotted gray line. For all the catalysts, the conversion increased by increasing the temperature, reaching the maximum at ca. 450–500 °C and decreasing afterward (Fig. 10a). As reported by several Authors, low temperatures were thermodynamically beneficial to the reaction but kinetically disadvantage due to the slow reaction rate [4,15,16,32]. Indeed, the 15Ni/GDC showed ca. 35% of CO₂ conversion at 300 °C, that increased to ca. 64% at 450 °C, remaining below the thermodynamic limits; this could be due to the great kinetic barrier for the full reduction of CO₂ (+4) to CH₄ (-4), an eight-electron process which required high activation energy [68]. Further increase in the reaction temperature led to a slightly decrease in the CO₂ conversion to ca. 60% at 600 °C, due to thermodynamic limit of methanation reaction and to the occurrence of RWGS reaction [32,68,69].

The variation on CH₄ concentration in the products mixture followed the same trend, increasing from ca. 9% at 300 °C to 17% at 450 °C and then decreasing to 8% at 600 °C (Fig. 10b). Apparently, this phenomenon was not thermodynamically consistent, because methane concentration should decrease by increasing temperature (Fig. 9) due to methanation equilibrium, suggesting that the methanation on the 15Ni/GDC catalyst at relative low temperatures ($T_{SET} < 450$ °C) was probably a kinetic controlled reaction [15]. Moreover, the CO byproduct concentration increased from 0.2 to 5.1% by increasing temperature from 400 to 600 °C due to the RWGS equilibrium.

The results showed that the increase in Ni loading positively affected the activity at all the investigated reaction temperature. The synthesized systems followed the methanation activity order: 50Ni/GDC > 30Ni/GDC > 15Ni/GDC. However, it is important to note that only slight variations in the catalytic activity were observed by increasing the Ni content. The 50Ni/GDC sample showed ca. 44% of CO₂ conversion at 300 °C, that reached the highest value of ca. 71% at 450 °C and then decreased to 65% at 600 °C (Fig. 10a). Despite several studies, the mechanism of CO₂ methanation is still under debate. Some authors suggested that CO₂ is first dissociated into CO either directly [70,71] or through carbonate and/or formate intermediates [72,73]. The formed CO would be then dissociated into C and O atoms on the

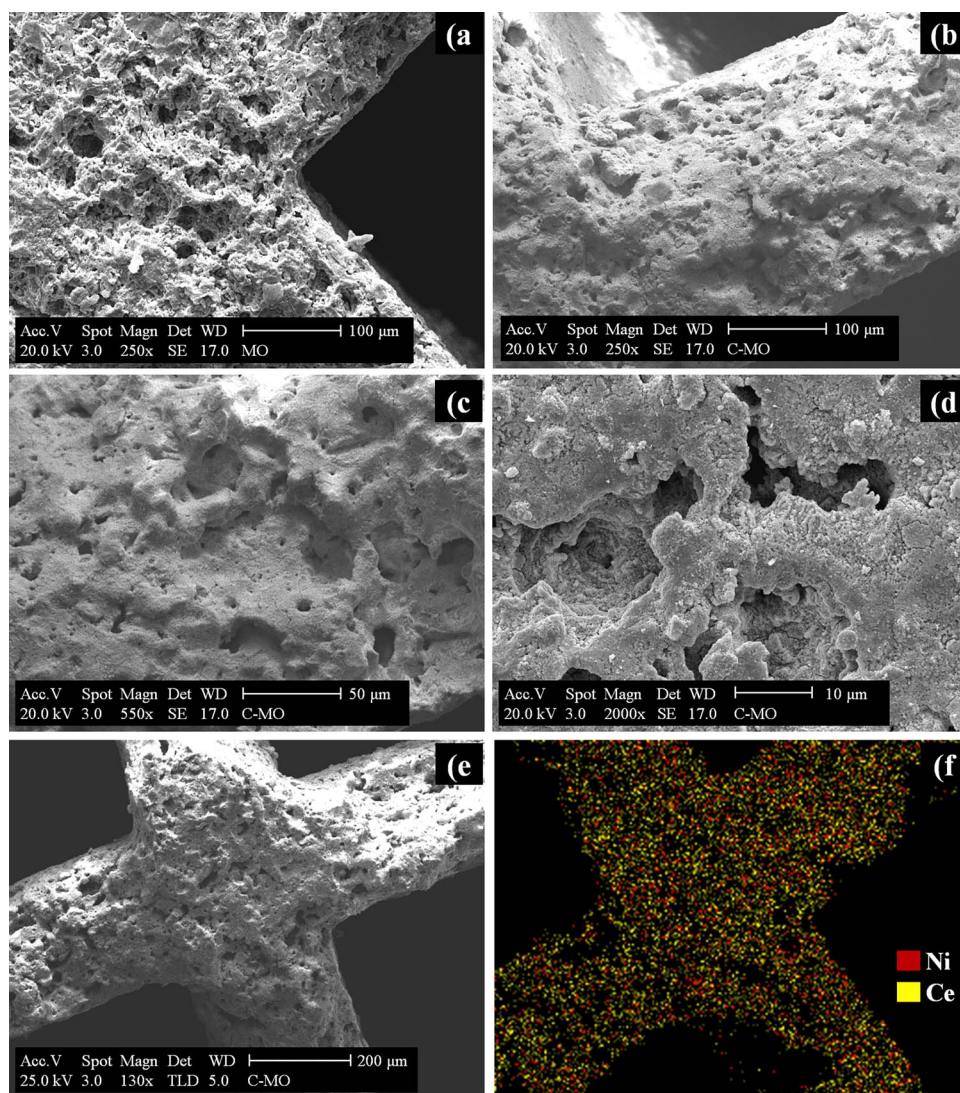


Fig. 7. SEM micrographs of the bare and coated (MO-B) monoliths: corner channel view of bare (a) and coated monoliths (b), inner channels view at different magnification of coated monoliths (c,d), frontal channel view (e) and corresponding EDX mapping (f) of coated monolith.

metal sites and further hydrogenated into methane [74]. Instead, other authors assumed that CO_2 methanation does not require CO as reaction intermediate but occurs via carbonates and formates which are directly hydrogenated into CH_4 [75–77]. However, for both mechanisms, the CO_2 adsorption on the basic sites of the catalyst is one of the key points to provide high methanation performance [14,18,32]. Thus, the enhanced activity of the 50Ni/GDC catalyst could be explained by the superior metal-to-support interactions and the increased number of moderate-basic sites, as confirmed by TPR and TPD characterization results. In addition, the presence of $\text{Ce}^{3+}/\text{Ce}^{4+}$ ion pairs and the

oxygen vacancies in the GDC support, as shown by the XPS analysis, could contribute to enhance the adsorption and activation of CO_2 [23]. On this basis, the 50Ni/GDC catalytic formulation was selected to prepare monolith-type structured systems.

3.2.2. Structured catalysts: effect of catalyst loading, reaction temperature and space velocity

To estimate the influence of external and internal mass transfer limitation, the Carberry (Ca) and Weisz-Prater (WP) numbers were calculated according to the following equations [78,79]:

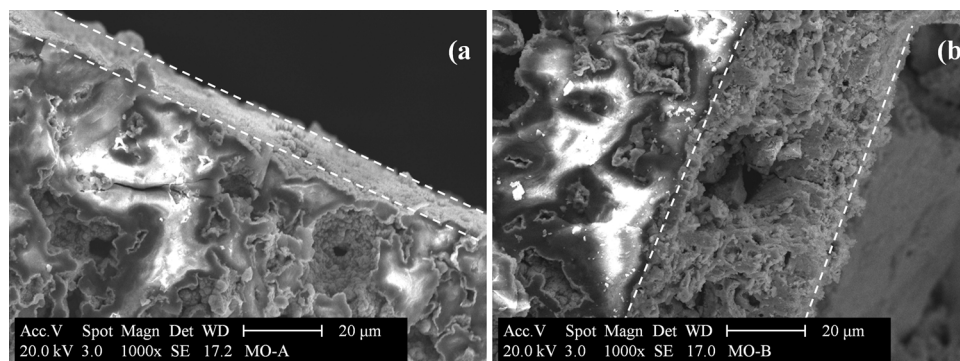


Fig. 8. SEM images of MO-A (a) and MO-B (b) coated layers.

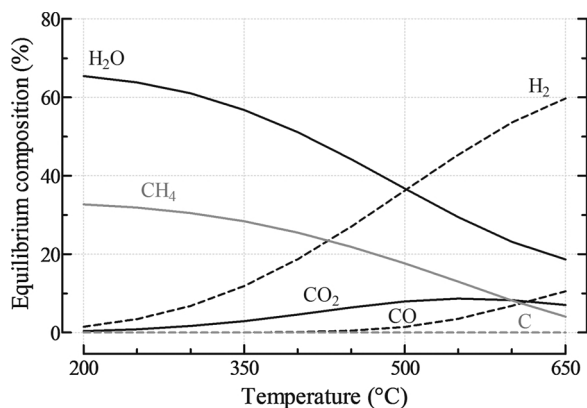


Fig. 9. Thermodynamic equilibrium analysis of CO₂ methanation: effect of temperature on products composition (H₂/CO₂ = 4/1 molar, P = 1 atm).

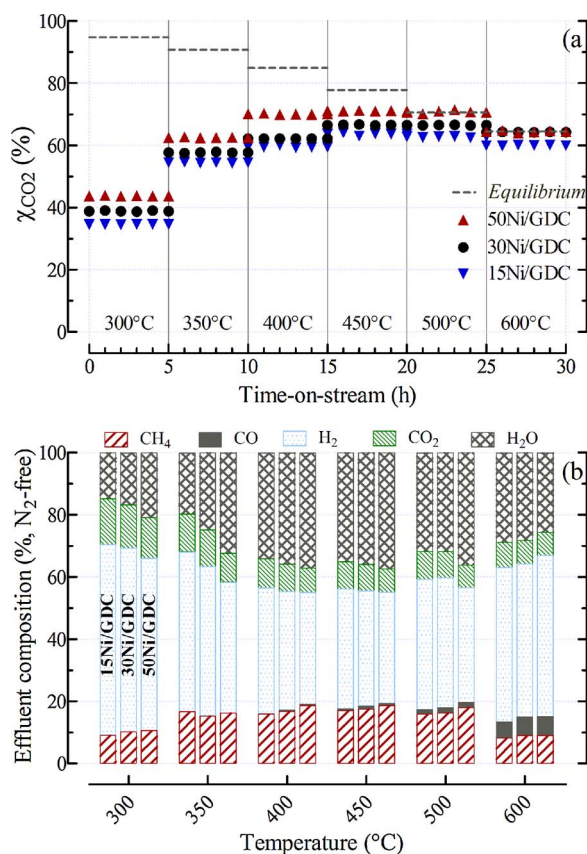


Fig. 10. CO₂ conversion (a) and effluent composition (b) over powder Ni/GDC catalysts. Influence of nickel content (15–50 wt.%) and set temperature (300–600 °C). Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, GHSV: 10,000 h⁻¹, total gas flow: 200 Nml/min, catalyst: 0.8 g/cm³.

$$Ca = \frac{r_{CO_2} \cdot \rho_{cat}}{k_G \cdot a_m \cdot C_{CO_2}} < \frac{0.05}{n} \quad (4)$$

$$WP = \frac{r_{CO_2} \cdot \rho_{cat} \cdot \delta_c^2}{D_{CO_2} \cdot C_{CO_2}} < \frac{n+1}{2} < 1 \quad (5)$$

where r_{CO_2} is the observed reaction rate with respect of CO₂ (8.51×10^{-5} kmol/kg s);

ρ_{cat} is the density of the catalytic layer (1.50×10^3 kg/m³);

k_G is the mass transfer coefficient (4.87×10^{-1} m/s), calculated by equation $k_G = S_h D_{CO_2} / d_h$, where S_h is the Sherwood number (2.99), D_{CO_2} is the diffusivity of CO₂ in gas phase (1.35×10^{-4} m²/s) and d_h is the hydraulic diameter (8.24×10^{-4} m);

a_m is the specific geometric surface area of monolith (2.52×10^3 m²/m³);

C_{CO_2} is the CO₂ concentration in gas mixture (7.44×10^{-3} kmol/m³);

δ_c is the characteristic coating layer dimension (4×10^{-5} m);

$D_{CO_2,e}$ is the effective diffusivity of CO₂ in the catalyst pores (7.73×10^{-7} m²/s), assuming catalyst porosity and tortuosity equal to 0.57 and 4, respectively [80].

Under the selected reaction conditions (i.e. P = 1 atm, T = 673 K, GHSV = 30,000 h⁻¹, MO-B structured catalyst), the calculated Ca and WP numbers for a first-order reaction rate (n = 1) were 0.014 and 0.035, respectively, confirming the absence of external and internal mass transfer limitations [78,79,81].

Fig. 11 shows the CO₂ conversion measured during methanation experiments over the structured catalysts with different catalyst loading (0.2 g/cm³ in MO-A and 0.5 g/cm³ in MO-B). Tests were carried out at atmospheric pressure, increasing the reaction temperature (T_{SET}) from 300 to 600 °C at GHSV of 10,000 (Fig. 11a), 30,000 (Fig. 11b) and 50,000 h⁻¹ (Fig. 11c), respectively. The temperature difference (ΔT) between the reaction temperature (T_{SET} = 350 °C) and the catalyst-bed temperature (T_{IN}, T_{CENTR}, T_{OUT}) is also depicted in Fig. 11d–f. The effluent compositions and the temperature profiles along the catalytic bed are summarized in Tables 5 and 6 for both MO-A and MO-B structured catalysts, respectively.

Poor activity was observed over the MO-A system at all the investigated GHSV (Fig. 11a–c). At GHSV of 10,000 h⁻¹, the CO₂ conversion increased from 7 to 66% by increasing temperature up to 400 °C, remaining lower than the equilibrium values due to kinetic limitations. At 450 °C, the CO₂ conversion (70%) approached the equilibrium value and slightly decreased to 63% by increasing the temperature up to 600 °C; this could be due to the thermodynamic limitation of methanation reaction and to the occurrence of the competitive RWGS reaction (Fig. 11a). Increasing the space velocity up to 30,000 (Fig. 11b) and 50,000 h⁻¹ (Fig. 11c), the CO₂ conversion decreased mainly due to the insufficient quantity of catalyst (0.2 g/cm³) especially at decreased contact time between reactants and catalyst. Analogously, the variation on CH₄ concentration in the products mixture showed the same trend (Table 5). These evidences could be related to the temperature profiles recorded along the catalytic bed (Fig. 11d–f, Table 5). The low-loaded MO-A monolith showed an almost flat temperature profile; as methanation is an exothermic reaction, this highlighted the poor reactivity of the sample. Indeed, the temperature values recorded along the catalytic bed were almost coincident with the set temperature, i.e. 355 (T_{IN}), 361 (T_{CENTR}) and 359 °C (T_{OUT}) at T_{SET} = 350 °C and GHSV = 10,000 h⁻¹ (Table 5). Increasing the GHSV up to 50,000 h⁻¹, lower temperatures were recorded along the catalytic bed (Table 5), due to the cooling effect induced by the reagents flow.

At relatively low space velocity, comparable activity was observed for the both structured catalysts (Fig. 11a), while higher activity was revealed for the MO-B system at GHSV of 30,000 and 50,000 h⁻¹ (Fig. 11b,c). This behavior, mainly due to the increased catalyst loading (0.5 g/cm³), is closely related to the temperature profiles recorded along the catalytic bed (Fig. 11d–f, Table 6). Although the high space velocity could create unfavorable conditions for the catalytic activity due to the short contact time, the activity of the MO-B system increased by increasing the GHSV from 10,000 to 50,000 h⁻¹ (Fig. 11), due to the higher temperatures recorded along the catalytic bed [82,83]. The high activity of the high-loaded MO-B system led to large amounts of released heat, induced by the strong exothermicity of CO₂ methanation, inevitably increasing the temperature along the catalytic bed [84]. Thus, the increasing of space velocity resulted in a much greater ΔT (temperature difference between the set reaction temperature and the catalytic bed temperatures) as reported in Fig. 11d–f. For example, MO-B catalyst showed 352 (T_{IN}), 365 (T_{CENTR}) and 346 °C (T_{OUT}) at T_{SET} = 350 °C and GHSV = 10,000 h⁻¹, corresponding to a maximum ΔT of 15 °C (Fig. 11d, Table 6). Increasing the GHSV to 50,000 h⁻¹,

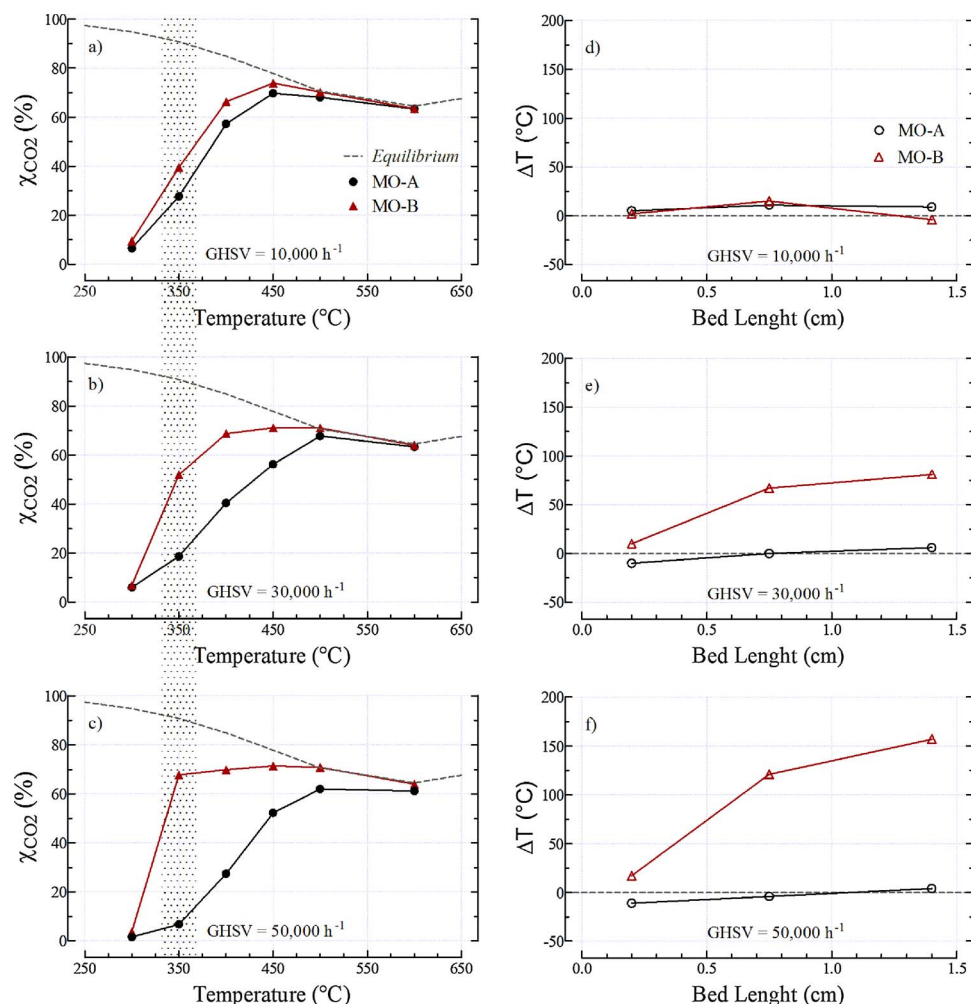


Fig. 11. CO₂ methanation activity over structured (MO-A and MO-B) catalysts. Influence of catalytic load (0.2–0.5 g/cm³) and set reaction temperature (300–600 °C) on CO₂ conversion at GHSV of 10,000 (a), 30,000 (b) and 50,000 h^{−1} (c). Temperature difference (ΔT) between the set temperature (T_{SET} = 350 °C) and the catalyst-bed temperature (T_{IN}, T_{CENTR}, T_{OUT}) at GHSV of 10,000 (d), 30,000 (e) and 50,000 h^{−1} (f). Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, total gas flow: 200–1000 Nml/min.

higher temperatures were recorded as T_{IN} (367 °C), T_{CENTR} (471 °C) and T_{OUT} (507 °C), with a maximum ΔT of 157 °C (Fig. 11f, Table 6). Under these conditions, CO₂ conversion increased from ca. 40% to 68% (Fig. 11a–c) by increasing GHSV from 10,000 to 50,000 h^{−1}.

Thus, the high activity of the high-load MO-B catalyst led to greater

amounts of reaction heat, that further increased by increasing the reagents flow. The released heat, transferred along the catalytic bed, affected the methanation performance, resulting in an almost flat CO₂ conversion profile in the temperature range 350–600 °C at GHSV of 50,000 h^{−1} (Fig. 11c). In addition, the CH₄ concentration increased

Table 5

Catalytic activity of structured (MO-A) catalyst for CO₂ methanation. Influence of space velocity (10,000–50,000 h^{−1}) and set temperature (300–600 °C) on effluent composition and temperature profile. Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, total gas flow: 200–1000 Nml/min, catalyst: 0.2 g/cm³.

GHSV (h ^{−1})	Set temperature (°C)	Effluent composition (% N ₂ -free)					Temperature profile (°C)		
		H ₂	CO ₂	CH ₄	H ₂ O	CO	T _{IN} (0.2 cm)	T _{CENTR} (0.75 cm)	T _{OUT} (1.4 cm)
10,000	300	77.4	18.6	1.4	2.6	0.0	300	301	301
	350	65.5	15.4	6.3	12.8	0.0	355	361	359
	400	47.4	10.9	15.8	25.3	0.6	412	418	408
	450	34.5	7.8	19.0	38.1	0.6	467	465	463
	500	36.3	7.8	19.4	35.5	1.0	520	514	492
	600	48.0	7.3	9.5	29.9	5.3	615	610	598
30,000	300	78.4	17.5	0.5	3.6	0.0	293	297	301
	350	75.4	16.4	2.1	5.9	0.2	340	350	356
	400	64.1	12.8	6.7	15.7	0.7	376	405	410
	450	51.5	10.4	12.6	23.2	2.3	416	466	460
	500	42.9	7.9	15.2	31.8	2.2	471	513	486
	600	46.5	7.1	11.5	29.9	5.0	594	612	583
50,000	300	78.2	20.0	0.3	1.5	0.0	291	295	300
	350	75.0	19.1	1.3	4.4	0.2	339	346	354
	400	65.3	15.7	5.0	12.5	1.5	376	403	413
	450	52.4	10.9	9.7	23.6	3.4	410	465	458
	500	47.3	9.0	11.8	27.9	4.0	455	515	480
	600	46.9	7.7	11.1	28.6	5.7	580	616	582

Table 6

Catalytic activity of structured (MO-B) catalyst for CO₂ methanation. Influence of space velocity (10,000–50,000 h⁻¹) and set temperature (300–600 °C) on effluent composition and temperature profile. Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, total gas flow: 200–1000 Nml/min, catalyst: 0.5 g/cm³.

GHSV (h ⁻¹)	Set temperature (°C)	Effluent composition (% , N ₂ -free)					Temperature profile (°C)		
		H ₂	CO ₂	CH ₄	H ₂ O	CO	T _{IN} (0.2 cm)	T _{CENTR} (0.75 cm)	T _{OUT} (1.4 cm)
10,000	300	76.6	16.0	2.3	5.1	0.0	301	305	301
	350	60.0	12.6	10.0	17.4	0.0	352	365	346
	400	37.6	8.1	19.0	35.3	0.0	407	420	379
	450	31.6	6.4	21.6	40.2	0.2	456	462	416
	500	35.2	7.0	19.3	37.8	0.7	510	509	466
	600	51.4	7.0	9.3	27.4	4.9	609	607	587
30,000	300	77.5	18.7	1.2	2.6	0.0	297	303	307
	350	53.7	11.1	11.7	23.5	0.0	360	417	431
	400	40.0	8.4	16.0	35.2	0.4	425	478	447
	450	35.6	7.4	17.6	38.2	1.2	479	512	460
	500	35.6	6.9	17.1	38.9	1.5	530	543	485
	600	47.1	7.5	11.0	29.5	4.9	624	624	588
50,000	300	77.7	19.8	0.8	1.7	0.0	298	306	316
	350	41.1	8.4	15.9	32.8	1.8	367	471	507
	400	35.7	7.5	17.1	37.7	2.0	443	536	512
	450	35.5	7.4	17.5	37.5	2.1	500	550	516
	500	39.3	7.7	15.6	34.8	2.6	544	585	527
	600	46.7	7.7	11.3	28.8	5.5	626	640	60

from 15.9% to 17.5% by increasing temperature from 350 to 450 °C, decreasing afterward to 11.3% at 600 °C, while CO concentration increased to 5.5% at 600 °C, mainly related to the influence of temperature on methanation and RWGS equilibriums (Table 6).

3.2.3. Powder and structured catalysts comparison

Powder and structured catalysts were tested by keeping unchanged the volume of catalytic bed (diameter, 1 cm, length, 1.5 cm). However, the MO-B system contained approximately 60% of the catalytic material employed in the packed bed reactor. As reported by Frey et al. [29], a better comparison between powder and monolith could be done considering the CH₄ productivity, expressed as hourly methane produced per unit weight of catalyst (L_{CH4}/g h). Fig. 12 compares the CH₄ productivity obtained with powder (50Ni/GDC) and structured (MO-B) catalysts as a function of reaction temperature (T_{SET} = 300–600 °C) and space velocity (GHSV = 10,000–50,000 h⁻¹).

As expected, MO-B monolith showed higher productivity values than the related 50Ni/GDC sample. High surface-to-volume ratio and good interphase mass transfer of the monolithic systems were able to ensure high methanation activity with low amount of catalytic phase [30,85]. Indeed, higher methane productivity was obtained with the MO-B catalyst (2.5 L_{CH4}/g h) than the 50Ni/GDC powder (1.6 L_{CH4}/g h). Moreover, the low pressure drops of structured system allowed operating at high space velocity [31,85]. Thus, the CH₄ productivity

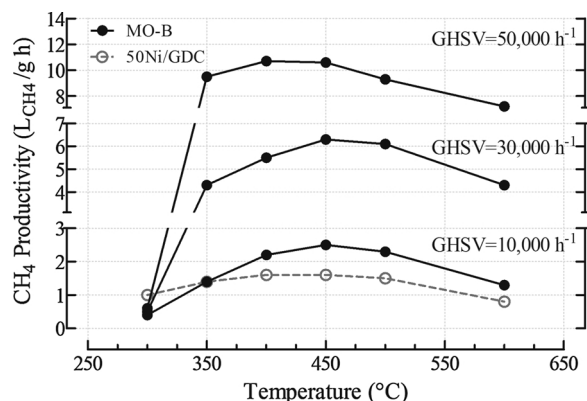


Fig. 12. Effect of set temperature (300–600 °C) and space velocity (10,000–50,000 h⁻¹) on CH₄ productivity values over powder (50Ni/GDC) and structured (MO-B) catalysts. Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, total gas flow: 200–1000 Nml/min.

increased by increasing the space velocity, reaching the highest value (10.7 L_{CH4}/g h) at 400 °C and 50,000 h⁻¹.

3.2.4. Stability test

A long-term test on the MO-B monolith-type structured systems was carried out at T_{SET} = 400 °C and GHSV = 30,000 h⁻¹, and the results are shown in Fig. 13. Stable performance, both in term of CO₂ conversion (68–69%) and effluent composition over 200 h of time-on-stream, were obtained. Moreover, almost stable temperatures were recorded along the catalytic bed, as reported in Fig. 13. Indeed, T_{IN}, T_{CENTR} and T_{OUT} slightly decreased from 422, 476 and 448 °C (revealed at the beginning of the test) to 417, 468 and 440 °C after 200 h, respectively, confirming the high stability of the studied system.

4. Conclusions

CO₂ methanation was investigated over powder and structured Ni/GDC (gadolinium-doped-ceria) catalysts at atmospheric pressure and H₂/CO₂ = 4 varying temperature (T_{SET} = 300–600 °C) and space

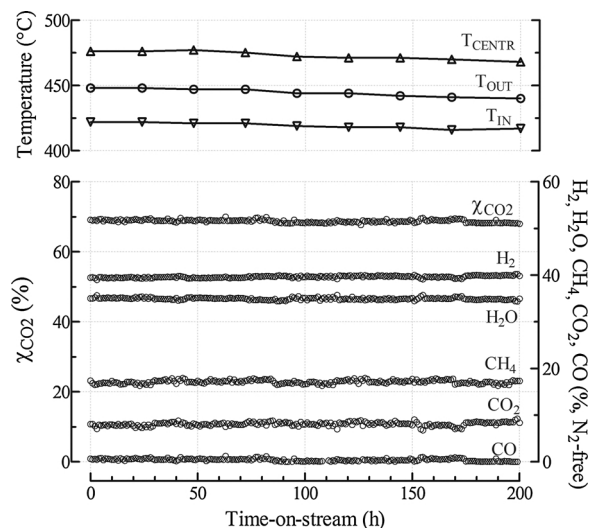


Fig. 13. CO₂ methanation stability of structured (MO-B) catalyst. CO₂ conversion and effluent composition as a function of time-on-stream. Reaction conditions: H₂/CO₂/N₂: 4/1/1 molar, set temperature: 400 °C, GHSV: 30,000 h⁻¹, total gas flow: 600 Nml/min, catalyst: 0.5 g/cm³.

velocity (GHSV = 10,000–50,000 h⁻¹). Powder systems were synthesized by solution combustion synthesis (SCS) varying the Ni content (15–50 wt.%). The activity increased by increasing the temperature while remaining below the thermodynamic limits, due to the kinetic limitations at relative low temperatures (300–450 °C). A slight decrease in activity was observed further increasing the temperature up to 600 °C, due to thermodynamic limitation of methanation reaction coupled with the occurrence of competitive RWGS equilibrium. Moreover, the activity increased by increasing the Ni loading due to enhanced metal-to-support interaction and moderate-basic sites, as confirmed by H₂-TPR and CO₂-TPD analysis. The presence of Ce³⁺/Ce⁴⁺ ion pairs and the surface oxygen vacancies in the GDC support also contributed to enhance the adsorption and activation of CO₂.

Monolith-type systems were prepared varying the amount of deposited catalyst on the structured support. It is worth mentioning that the catalyst preparation appeared fully reproducible. Very uniform, thin (10–40 µm) and high-resistance catalytic layers were *in situ* deposited on the cordierite monoliths by the SCS method. Low-loaded monolithic catalyst (0.2 g/cm³) showed poor activity, especially at decreased contact time between reactants and catalyst, due to the insufficient amount of catalytic layer. Relative higher activity was recognized using the high-load monolith (0.5 g/cm³), closely connected to the temperature profiles recorded on the catalytic bed. High surface-to-volume ratio, good interphase mass transfer and low pressure drops of the structured catalyst led to high methane productivity per unit weight of catalyst. The highest CH₄ productivity of 10.7 L_{CH4}/g h, was obtained at 400 °C and 50,000 h⁻¹. Promising long-term stability was observed over 200 h of time-on-stream.

The results presented in this paper provided a platform to further optimization studies of structured catalysts for CO₂ methanation in power-to-gas applications.

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